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Restrictions and Limitations for the Design of a Steam Generator for a Coal-fired Oxyfuel Power Plant with Circulating Fluidised Bed Combustion

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Abstract

As general restrictions for the design of Oxyfuel pulverised coal-fired (PC) steam generators are commonly known, the purpose of this work is to show general restrictions for the design of an Oxyfuel coal-fired steam generator using a circulating fluidised bed combustor (CFBC) with external heat exchangers (EHEs). For the CFBC restrictions result on the one hand out of the used fuels composition. On the other hand certain restrictions out of state-of-the-art and steam generator geometries have to be considered within the design.

Keywords: Oxyfuel; steam generator; fluidised bed; CFBC; CO_2 capture

1. Introduction

The Oxyfuel process is one of the three main paths – Post-, Pre-Combustion and Oxyfuel – to avoid CO_2-emissions of conventional power plants by carbon capture and storage. The applied process bases on the exclusion of nitrogen from the process. Instead of an N_2/O_2 –atmosphere the combustion takes place in an O_2/CO_2-atmosphere, which is mostly realised by an oxygen supply with a cryogenic air separation unit (ASU) and a flue gas (FG) recirculation, which is necessary to deliver a heat sink for the otherwise extraordinary high flame temperatures. The PC-fired option for the Oxyfuel process is well examined. To realise a manageable temperature on the flue-gas-side (furnace exit temperature, limited by ash softening

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temperature) and the water-steam-side (maximal steam temperature at furnace outlet in the membrane wall, limited by allowable material temperature), about 2/3 or more of the flue gas has to be recirculated [1], [2]. The motivation to use a CFBC firing system is to reduce this high flue gas recirculation. In contrast to Oxyfuel PC-fired systems with the heat sink of recirculated flue gas only, the CFBC also has the option to substitute a part of recirculated flue gas by cooled solids. This delivers one more degree of freedom for the steam generator design and by this the possibility to reduce the necessary flue gas recirculation, although the temperature at the outlet of the combustion chamber (CC) is significantly lower than in a PC unit. The flue gas recirculation is defined as the sum of mass flows recirculated to fluidise the CC, the EHEs and the loop-seals, related to the flue gas mass flow leaving the convective heat exchangers (CHE). A schematic comparison of the PC and CFBC firing systems can be obtained out of Figure 1. An exact value for the recirculation rate for the CFBC cannot be mentioned here, as it strongly depends on assumptions made for the process, especially on used fuels.

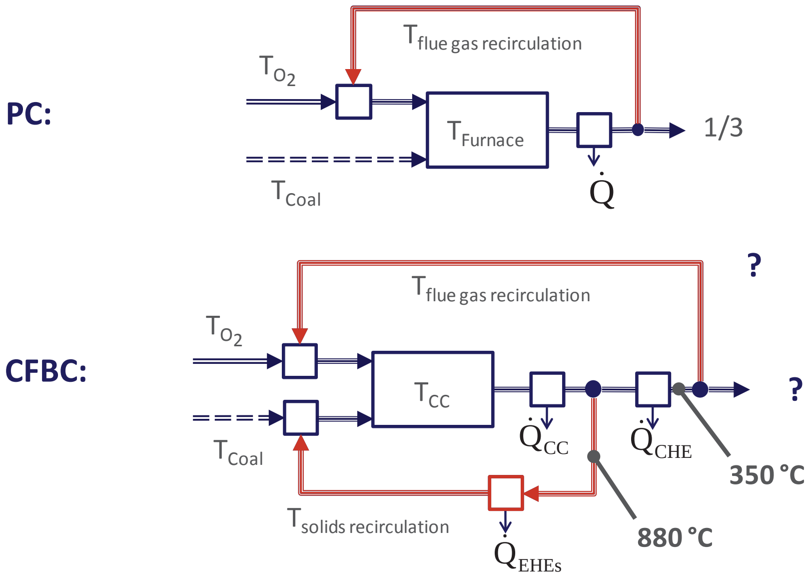

Figure 1: Comparison of PC and CFBC firing for Oxyfuel conditions

2. The Oxyfuel process with CFBC

As can be seen in Figure 1 the Oxyfuel process for a CFBC differs from the PC option especially by the use of cooled solids as a heat sink. To be able to use this heat sink, the hot particles separated in the cyclones have to be cooled down in EHEs. As these have a certain fluidisation demand and a higher pressure loss than the CC, a process scheme like shown in Figure 2 for the Oxyfuel CFBC is considered in this work.

For the ASU a configuration with adiabatic compression is chosen. The oxygen leaving the ASU has a temperature of about 180 °C and a purity of 95 vol.-%. This lower purity of 95 vol.-% shows a significant advantage concerning the electrical net efficiency of the overall process in comparison to much higher oxygen purities [3].

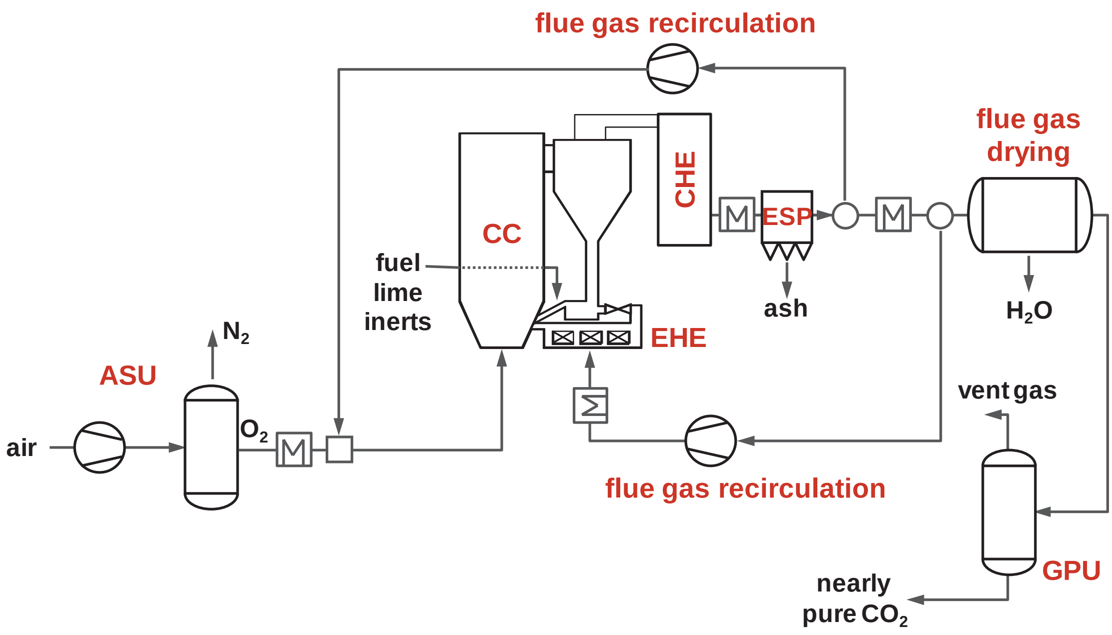

Figure 2: Simplified process scheme of an Oxyfuel CFBC process

The oxygen leaving the ASU is further preheated up to approx. 315 °C by flue gas upstream the electrostatic precipitator (ESP) and mixed with a hot recirculation (approx. 245 °C), taken directly downstream of the ESP. The mixed oxidant is fed into the CC as a primary oxidant via nozzle floor or as a secondary oxidant above the refractory lining. The resulting flue gas and the entrained solids out of the CC are led to the cyclones where most of the solids are separated. While the flue gas enters the CHE to transfer sensitive heat to the water-steam-side, the separated solids first enter a loop-seal, where a fraction is forwarded to the EHEs and the rest is directly returned to the CC. After cooling in the EHEs solids are mixed with fuel, additional inert material and sorbents, like CaO or $CaCO_3$, in the return leg and afterwards return to the CC.

Beside for preheating the oxygen flue gas leaving the CHE is used for preheating the recirculated flue gas fluidising the EHEs and the loop-seals (approx. 315 °C). As there is still enough flue gas heat available on a high temperature level an HP-bypass on the water-steam-side is considered, before the flue gas enters the ESP. A part of the flue gas is directly recirculated for oxidant mixing, while the remaining part of the flue gas is further cooled down to a lower temperature level with an LP-bypass, to realise a compression of further recirculated flue gas to about 1.5 bar abs for the EHEs and the loop-seal fluidisation. As the pressure drop of the EHEs is factor three to four higher compared to the CC, the EHEs have to be fluidised by a second recirculation. Due to the pressure difference for fluidisation and the larger compression the recirculation for the EHEs fluidisation needs a branch point further downstream the process, to have a lower flue gas temperature level and by this a possibility to compress the flue gas.

Subsequently the flue gas enters a flue gas drying system, to withdraw rest water, before entering the gas processing unit (GPU). Inside the GPU the flue gas stream is separated into a CO_2-rich and a CO_2-lean stream, where the CO_2-lean stream leaves the process as vent gas of the GPU, while the rectified and liquefied CO_2-rich stream can be compressed and further transported to a storage site.

2.1. The oxygen ratio for Oxyfuel processes

One of the most important parameters for the Oxyfuel process is the oxygen ratio. In comparison to the air case, where all oxygen entering the steam generator can be exactly determined by a residual oxygen measurement in the flue gas, this is not possible for the Oxyfuel process. Because of the recirculation of flue gas there is an extra input of oxygen into the steam generator, which is delivered by the residual oxygen in the flue gas. This extra oxygen input is significant, though the oxygen content in the flue gas is relatively low. It can be seen as difference between the local and the global oxygen ratio in Figure 3.

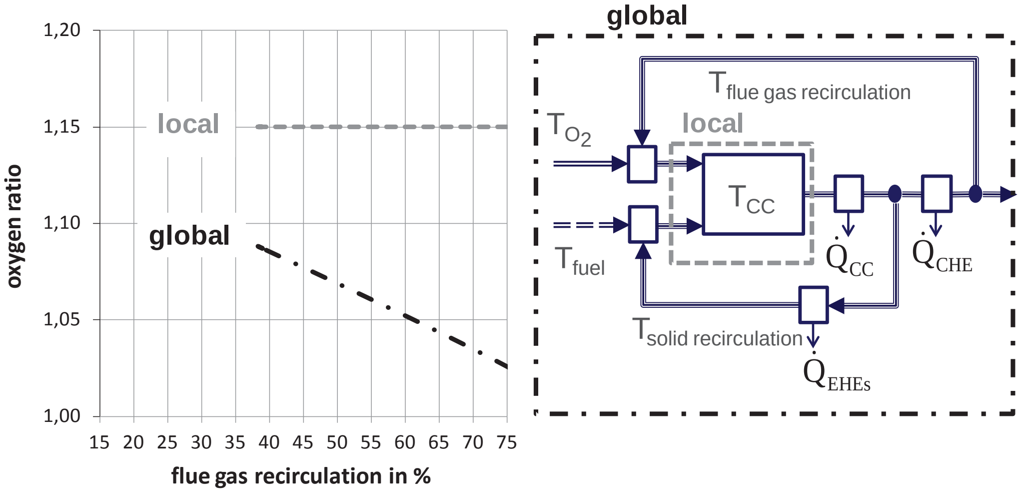

Figure 3: Different approaches for the oxygen ratio in the Oxyfuel Process

For the Oxyfuel process there are two different definitions for the oxygen ratio. On the one hand a global oxygen ratio of the overall process (dash dotted in Figure 3) can be defined; on the other hand a local oxygen ratio for the steam generator (dotted in Figure 3) can be defined. For PC fired boilers the fulfilment of a local oxygen content of 1.15 at full load is strictly recommended [4]. Although this ratio might be reduced for CFBC Oxyfuel the same ration is chosen here (see left hand side diagram of Figure 3). Keeping the local oxygen ratio at this constant level while increasing the flue gas recirculation ratio leads to a decrease of the global oxygen ratio as more of the residual oxygen of the flue gas is recirculated. As the global oxygen ratio decreases, less oxygen has to be supplied by the ASU.

3. Modelling of the process

The modelling of flue gas and water-steam-side of the Oxyfuel process is done with EBSILON®*Professional*. Because the solid side is not implemented in the software as detailed as needed, certain calculations were added with so called *Kernelscripts*, see also 3.1.

Results shown in 4 base on a greenfield power plant, which is orientated at Lagisza power plant in terms of steam parameters and gross power output [5]. Lagisza was chosen as reference in these points, as it is one of the newest CFBC power plants and by this represents state-of-the-art, while involving the highest live steam parameters realised with a CFBC so far. The preheating train consists of three LP preheaters, a feed water tank and three HP preheaters, while the condenser pressure is 45 mbar. For the Oxyfuel process additional cooling in comparison to a conventional power plant becomes necessary for the ASU and the GPU. The GPU is simulated as isobaric CO_2 condensation with two-stage external cooling-system, for further details and results see also [6]. The chosen coal for simulations is a South African hard coal with a lower heating value of about 25.1 MJ/kg. This coal was chosen to have a direct comparison of PC and CFBC performance, see also [7]. Further such coals are still used in CFBC boilers and CFBC boilers are still designed for such coals [8]. The turbine efficiencies taken into consideration for modelling base on state-of-the-art isentropic efficiencies for HP, IP and LP turbines, while in the last stages of the LP turbine influences by droplets of wet steam (Baumann Correlation) are considered.

Table 1: Characteristics of the power plant model

Gross output	460 MW_{el}
Live steam temperature	560 °C
Live steam pressure	275 bar
Reheat temperature	580 °C
Condenser pressure	45 mbar
FBC temperature	880 °C
EHE outlet temperature	550 °C
Flue gas outlet Economiser	340 °C
South African Hard Coal (LHV)	25.1 MJ/kg
Oxygen ratio (local)	1,15
Empty tube velocity (FBC)	5 m/s
Air ingress	0,5 %
desulphurisation efficiency	95 %
Oxygen purity	95 vol.-%
ASU specific demand	236 kWh/t_{O2}
CO_2 capture rate	90 %

In the CC, the loop-seals and the EHEs no air ingress is considered, because these parts of the plant are operated at higher than ambient pressure. The welding of the CHE path should be gas-tight, so no air ingress is considered here, too – this is in contrast to PC-fired boilers, where an additional air ingress has to be considered e.g. for the gap between burner and furnace or for burner cooling etc. [9]. Therefore in the CFBC Oxyfuel process air ingress is assumed only in the area of the high-temperature ESP where an air ingress of 0.5 %, related to the volume flow under standard condition, is considered. All chosen boundary conditions are in agreement with data from manufacturing and supplying companies. The most important boundary conditions for the model are listed in Table 1.

For the desulphurisation many different efficiencies, depending on the coal and the Ca/S-ratio have been shown e.g. on the 2nd *Oxyfuel Combustion Conference* in 2011, see also [10] and [11]. Because it is not sure whether a direct sulphatisation or an indirect sulphatisation of the added limestone will occur under Oxyfuel conditions, it is assumed that no secondary measures have to be installed into the process and a desulphurisation efficiency of 95 % can be realised.

3.1. Modelling of CC, EHE and loop-seals

For the solid-side of the model different mathematical approaches were programmed to simulate all key components – CC, EHEs and loop-seals. As all approaches base on a certain terminal velocity of particles and their entrainment, for simulations a monodisperse particle size of 150 μm is chosen. This value is a result of several studies about particle size distributions during operation of CFBC [12], [13].

As no heat transfer has to be implemented for loop-seals the approach by Basu was chosen for design purposes [14]. For the CC as well as the solid entrainment, the heat transfer for industrial scale steam generators is implemented in the model. For the solid entrainment the approach of Kunii and Levenspiel [15] was chosen, because the analyses of several operating power plants have also been modelled with this approach with good agreement to measured data [12], [16]. The heat transfer in industrial scale power plants was analysed by Leckner and Breitholtz. As the results were compared with several other studies, the approach mentioned in [17] was chosen. To build the CC as small as possible and to maximise the heat withdrawal in the furnace, additional heat transfer surfaces, so called wing-walls were included in the steam generators design. For these the approach of Leckner and Breitholtz is assumed, too. Further heating surfaces, so called platen superheaters, are considered in the design as well. As these show a certain different heat transfer the approach made by Basu, mentioned in [18], was chosen for modelling.

The EHEs become more important in comparison to the air case, as the CC is supposed to be as small as possible and by this heat is transferred in the EHEs from solid to water-steam-side to a greater extent. The EHEs are modelled as a bubbling fluidised bed with an approach by Werther [19] as this fits best to the heat transfer model of Martin, mentioned in [20] and [21]. All EHEs consist of three chambers. In the first chamber the solids are only held in movement and the solid flow entering the following chamber is homogenised. The other two chambers are realised as heat transfer sections of the EHEs. This setup is chosen as a consequence of erosion problems [22]. Moreover from a design point of view the EHEs size is restricted, as solid transport by the loop-seals and the stand pipe to the EHEs has to be ensured.

4. Results

The following results are for an Oxyfuel CFBC under full-load conditions. The focus is drawn on limitations of the process and differences of a CFBC designed for an air and an Oxyfuel case.

First it has to be mentioned, that a constant flue gas velocity in the CC leads to a larger cross-sectional area for the CC with an increase of flue gas recirculation. This is based on higher mass and volume flows inside the CC. With more available space inside the CC additional platen superheaters and wing-walls can be realised. Moreover further cyclones, loops-seals and EHEs can be placed around the CC. The cross-sectional area increase leads to a discontinuity in the transferred heat for the CC, as with an increased cross-sectional area enough space for another wing wall is available. This can be seen in Figure 4 (a) in which the distribution of transferred heat in CHE, CC and EHEs is plotted against flue gas recirculation. Additionally the heat distribution for an air-case is shown in the plot.

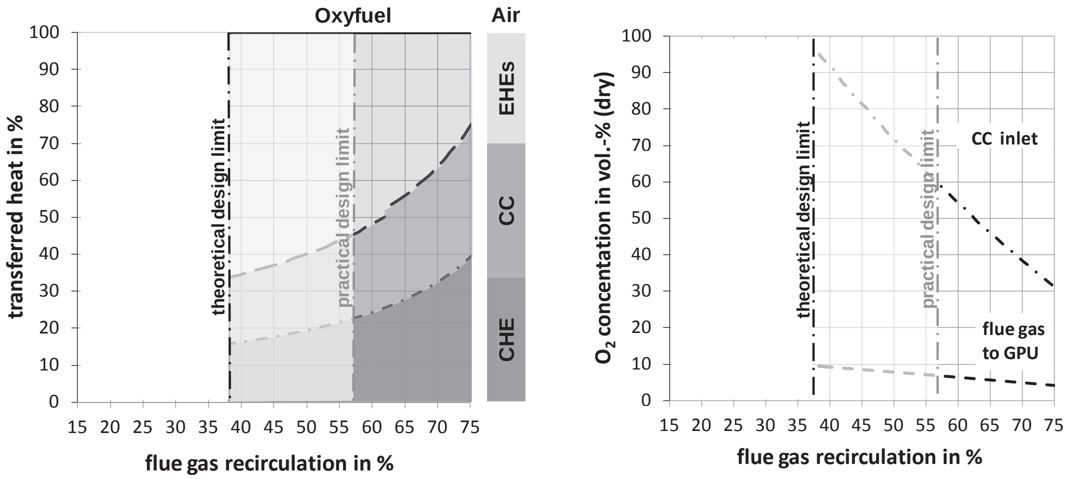

Figure 4: (a) Transferred heat in the CHE, CC and EHEs for a variation of the flue gas recirculation of an Oxyfuel CFBC on the right side are given the explanations and the heat distribution for the air case; (b) The dry O_2 concentration of the flue gas and the oxidant at the entry of the CC

For the air case parts of transferred heat in CHE (32 %), EHEs (33 %) and CC (35 %) are nearly equal. This is comparable to a flue gas recirculation of about 70-75 %.These parts vary with a variation of the flue gas recirculation at constant CC and EHEs outlet temperatures. In general the amount of heat transferred in the CHE is decreasing with a decreasing recirculation rate. This is based on less available flue gas mass flow transferring heat in the CHE at a constant temperature difference between inlet (880 °C) and outlet (340 °C). The amount of heat transferred in the CC decreases as well, as the cross-sectional area decreases and by this available wall surface and additional surfaces are reduced. Because both CC and CHE decrease in the amount of transferred heat, more heat has to be transferred to the water-steam-side inside the EHEs.

As can be seen in Figure 4 (a) the flue gas recirculation cannot be reduced to zero, because the EHEs and the loop-seals always have a certain fluidisation demand. This is the reason for a minimum EHEs fluidising gas demand shown in the figure, which is for set assumptions at about 38 % flue gas recirculation. This value represents the absolute theoretical design limit and the minimal flue gas recirculation of the process. It is highlighted in Figure 4 (a) with black lettering.

For the design of the CC 38 % flue gas recirculation is not the only restriction. Considering a certain velocity above the nozzle floor in the CC, depending on the necessary angle of the hopper caused by the used type of coal and its volatiles, another limit for the design is set. As these values are strongly depending on the manufacturers know-how, it is assumed, that half of the CC cross-section is needed as space for the nozzle floor. This leads to a restriction of the process at about 45 % flue gas recirculation. This value is still not the final limitation for the process concerning state-of-the-art. The minimal flue gas recirculation for the simulated process set-up and by this the practical design limit is at about 57 %, highlighted in Figure 4 (a) with grey lettering. This limitation is based on the size and design of EHEs. As these have a maximal solid flow rate and as the space around the CC is limited as well, the minimal flue gas recirculation cannot be lower than this value for chosen assumptions for feasibility of the process. The boundary set by the last mentioned limitation can be further restricted with the occurrence of problems during combustion, these still have to be further analysed and are neglected in these analyses. But - for the record - the realised oxidant staging, like done for the air case with about 60 % secondary air, see also [13], will not be possible under these conditions. A staging of the oxidant will be about 70 % primary oxidant and 30 % secondary oxidant for the Oxyfuel case, concerning a flue gas recirculation of about 70 %.

Taking a look at the heat distribution for an air and an Oxyfuel case shows, that there has to be an increase in the size or number of the EHEs. These have to transfer about two third more heat compared to the air case for the practical design limit (57 %), and by this have to transfer more than 50 % of the total heat. The reason for this is a decrease of heat transferred in the CHE (-1/3) and the CC(-1/3).

In Figure 4 (b) the dry O_2 concentration of the flue gas and the dry O_2 concentration of the oxidant at the entry of the CC are plotted against flue gas recirculation. In general a reduced flue gas recirculation leads to higher O_2 concentrations at the inlet of the CC as well as to higher O_2 concentrations in the flue gas. The reason for the increased residual oxygen content in the flue gas is that the same amount of oxygen, given by the local oxygen ratio of 1.15, is mixed with less CO_2 due to a lower recirculation rate. The O_2 concentration at the nozzle floor is increasing with less flue gas recirculation due to the same reason and additionally because a major part of the flue gas recirculation is needed for the fluidisation of the EHEs and the loop-seals. This effect is intensified by a bigger fluidisation demand of the EHEs with less flue gas recirculation, because more heat has to be transferred in the EHEs.

The design limit values mentioned above, are also shown in the plot of Figure 4 (b). For a flue gas recirculation of 38 %, the O_2 oxidant concentration is about 95 vol.-% and by this the same concentration as coming from the ASU and thus shows, that all recirculated flue gas is needed for fluidisation of the EHEs and the loop-seals. Such high concentrations of oxygen at the CC hopper will lead to problems during combustion. Hot spots in the dense bed of the CC hopper will occur and by this temperatures, higher than ash softening temperature at these hot spots will cause sintering and agglomeration of the ash. A boiler shutdown will be the consequence. To avoid this risk lower oxygen concentrations at the CC nozzle floor are necessary to ensure a secure operation of the CFBC.

The limitation of 57 % flue gas recirculation shows a much lower O_2 concentration for the oxidant entering the nozzle floor in comparison to 38 % flue gas recirculation. The O_2 concentration is about 60 vol.-% (dry) and thus still about three times higher than for the conventional air case, which might still cause problems during operation of the plant caused by sintering and agglomeration. Especially the restricted oxidant staging might lead to higher bed temperatures and by this to worse in-situ desulphurisation and higher NOx-emissions compared to the air case.

5. Summary and outlook

The motivation to use an Oxyfuel process with CFBC instead of an Oxyfuel process with PC firing is to reduce the necessary flue gas recirculation. The analysis of the Oxyfuel CFBC is done from a theoretical point of view. To be able to estimate whether the process is feasible or not, deeper analysis of the combustion behaviour under Oxyfuel conditions in a CFBC have to be undertaken. Especially such high O_2 concentrations of 60 vol.-% (dry) or even higher in the oxidant should be analysed, to receive a certainty of feasibility, as this might lead to agglomeration and boiler shutdowns due to hot spots in the bed. Anyway from a theoretical point of view a feasibility seems to be given and an even lower flue gas recirculation might be realised, when the steam generator is designed in more detail.

Furthermore desulphurisation under Oxyfuel conditions needs to be analysed deeper, as only few knowledge and reliable results are available, due to different behaviour of different limestones and coals analysed so far. Depending on the success of desulphurisation and oxidant staging the success of the CFBC under Oxyfuel conditions will show whether the CFBC can compete against a PC fired plant in terms of net efficiency and availability.

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